



1 However, increasingly tighter regulations on contaminant in fuels, particularly  
2 sulfur and aromatics, have forced many refiners to hydrorefine most and often  
3 all, of the fuel products. To meet the more stringent requirements for low  
4 sulfur diesel, refiners have added naphtha hydrotreaters for removing sulfur  
5 and nitrogen compounds from at least some of the refinery streams which go  
6 to make up the gasoline pool. In response to the more stringent requirements  
7 for clean diesel fuels, refiners have added diesel hydrotreaters for making the  
8 low sulfur, low aromatics diesel which are now preferred, and often required.  
9 More refiners are building hydrocrackers due to their ability to produce high  
10 quality low sulfur fuels. The light gaseous products processed in a refinery  
11 are generally treated to remove H<sub>2</sub>S and others sulfur containing components  
12 prior to use of the gaseous products for energy, as petrochemical feedstocks,  
13 as reforming feedstocks for making synthesis gas, or as building blocks for  
14 turning the gaseous products into higher molecular weight products.

15  
16 Thus, in response to these tightening regulations, refiners have constructed  
17 separate hydroprocessing units to upgrade each of the fuel streams produced  
18 in the refinery. The net effect is a large number of similar processing units,  
19 each handling a separate stream, requiring additional tankage and operators.  
20 Specific streams are alternatively heated for reaction or fractionation, and  
21 then cooled for separation and storage. Multiple reaction systems requires  
22 multiple hydrogen supply, pressurization and distribution systems. It is  
23 desirable to have a process for hydroprocessing the entire crude oil into  
24 useful low aromatic, low sulfur products while significantly reducing the  
25 number of refinery processing steps and processing equipment required to  
26 convert the crude to useful products. Such a process is the subject of the  
27 present invention.

28  
29 In U.S. Patent No. 5,009,768, a complete crude or the atmospheric and  
30 vacuum residues thereof mixed with vacuum gas oils is demetallized and the  
31 demetallized product hydrotreated for hydrodenitrogenation and  
32 hydroconversion. In U.S. Patent No. 5,382,349, a heavy hydrocarbon oil is  
33 hydrotreated, the hydrotreated oil distilled and a vacuum residue thermally  
34 hydrocracked in a slurry bed. U.S. Patent No. 5,851,381 provides a method

1 of refining crude oil by distillation and desulfurization. In the method, a  
2 naphtha fraction is separated from crude oil by distillation, with the remaining  
3 residual fraction after the naphtha fraction has been removed from the crude  
4 oil being hydrodesulfurized and the hydrodesulfurized fraction separated into  
5 further fractions, first in a high pressure separator and then by atmospheric  
6 distillation. A residue is further upgraded in a residue fluid catalytic cracking  
7 process.

### 8 9 SUMMARY OF THE INVENTION

10  
11 In the present process, a crude oil feed is desulfurized and processed  
12 (hydrotreated and hydrocracked) to form low sulfur, low aromatic fuels in an  
13 integrated unit, with a single hydrogen supply and recovery loop, with minimal  
14 cooling of intermediate products, and without tank storage of intermediate  
15 products. The integrated unit comprises a series of catalytic reaction zones,  
16 each containing a single catalyst or a layered catalyst system selected for a  
17 particular application, whether it be desulfurization of a crude feed,  
18 hydrocracking a gas oil stream or hydrotreating a particular stream to reduce  
19 the aromatic and/or sulfur content of the stream to low levels. Flash  
20 separation of reaction products exiting a particular catalytic reaction zone is  
21 tailored to isolate hydrogen with minimal heat exchange beyond that required  
22 to prepare the reaction products for the next processing step.

23  
24 In the present invention, a crude oil feed is passed directly to a crude  
25 desulfurization unit for desulfurization. The crude oil feed may be desalted  
26 and volatile materials removed prior to desulfurization, but a substantial  
27 portion of the crude oil feed is subjected to desulfurization in a desulfurization  
28 reaction zone. A number of reactions is expected to occur during the  
29 desulfurization process. Portions of the crude oil feed which contain  
30 metal-containing components will be at least partially demetallized during the  
31 desulfurization process. Likewise, nitrogen and oxygen are removed, along  
32 with sulfur, during the desulfurization process. While the amount of cracked  
33 products produced during desulfurization will be relatively small, some amount

1 of larger molecules will be cracked to lower molecular weight products during  
2 the desulfurization process.

3

4 The desulfurized crude oil temperature is adjusted for fractionation, and a gas  
5 oil fraction isolated. The gas oil fraction is available for use directly as a fuel.  
6 Preferably, the gas oil fraction is further hydrotreated for additional sulfur,  
7 nitrogen and/or aromatic removal. Yields of desirable fuel products are  
8 increased in the present process when the desulfurized crude oil product is  
9 fractionated, preferably in a multi-stage fractionation zone having atmospheric  
10 and vacuum distillation columns. Products from multi-stage distillation include  
11 a light gas oil fraction, a vacuum gas oil fraction and a residual fraction. The  
12 light gas oil fraction, generally having a normal boiling of less than 700°F, may  
13 be used directly as a fuel, or further hydroconverted for improved fuel  
14 properties. The vacuum gas oil fraction is hydrocracked to increase the fuel  
15 yield in the present process and to further improve fuel properties. Single or  
16 multi-stage hydrocracking reactors may be employed. The hydrocracked  
17 products includes at least one low sulfur fuel product, which may be isolated  
18 from a step of distilling the hydrocracked products.

19

20 Accordingly, a process is provided for hydridesulfurizing a crude oil feed in a  
21 crude desulfurization unit, separating the desulfurized crude oil and isolating a  
22 light gas oil fraction, a vacuum gas oil fraction and a residual fraction,  
23 hydrocracking the vacuum gas oil to form at least one low sulfur fuel product;  
24 and hydrotreating the light gas oil fraction. This entire integrated process may  
25 be conducted without using tank storage of intermediate products, such as a  
26 desulfurized crude oil, a light gas oil fraction, and a vacuum gas oil fraction.  
27 Further, with no required tank storage of intermediate products, the preferred  
28 process can be conducted without cooling of the intermediate products, thus  
29 reducing the operating cost of the process. In a further cost savings, the  
30 hydroconversion steps of the present process, including crude desulfurization,  
31 hydrocracking and hydrotreating, are suitably conducted using a single  
32 hydrogen supply loop, thus further reducing the capital and operating cost of  
33 the process.

1 The present invention provides an integrated refining system for processing a  
2 whole crude, or a substantial portion of a whole crude, into a full range of  
3 product materials at high selectivities and high yields of the desired products.  
4 The integrated process of the present invention further provides a series of  
5 reaction zones, containing catalysts of varying pore volume, for successively  
6 converting progressively lighter and cleaner products in the production of fuel  
7 products. The integrated process further provides an method for isolating,  
8 purifying and providing hydrogen to the various conversion reaction zones  
9 through the use of a single hydrogen isolation and pressurization unit. Among  
10 other factors, the present invention is based on an improved understanding of  
11 hydroconversion processes, permitting more efficient use of a combination of  
12 units for reaction, for product isolation, for hydrogen isolation and recycle, and  
13 for energy usage in the preparation of fuels from a crude feed. In the process,  
14 a wide range of fuel oil products can be safely prepared with a small number  
15 of reaction vessels and product recovery vessels, and with a minimum  
16 number of supporting vessels, for handling hydrogen and intermediate  
17 products, and employing a minimum number of operators. In effect, the  
18 present invention is based on the novel combination of crude desulfurization  
19 tailored to a wide boiling range feed, followed by distillation to form a few  
20 distillate streams, and bulk upgrading in an integrated  
21 hydrocracking/hydrotreating process to form a wide range of useful fuel and  
22 lubricating oil base stock products. The present process provides an efficient  
23 and less costly alternative to the conventional refinery practice of separating a  
24 crude oil feed into a number of distillate and residuum fractions, each of which  
25 are processed individually in similar but separate upgrading processes.

## DESCRIPTION OF THE FIGURES

Figure 1 discloses a crude oil desulfurization process which comprises the following steps:

- a) hydrodesulfurizing a crude oil feed in a crude desulfurization unit;
- b) separating the desulfurized crude oil and recovering a light gas oil fraction, a vacuum gas oil fraction and a vacuum residuum fraction;
- c) hydrocracking the vacuum gas oil to form at least one low sulfur fuel product; and
- d) hydrotreating the light gas oil fraction.

Figure 2 discloses a crude oil desulfurization process which comprises the following steps:

- a) hydrodesulfurizing a crude oil feed;
- b) separating the desulfurized crude oil and recovering at least a light gas oil fraction, a vacuum gas oil fraction and a residual fraction;
- c) hydrocracking the vacuum gas oil in a first hydrocracking reaction zone to reduce the sulfur content and the nitrogen content therefrom and to produce a low sulfur gas oil product;
- d) hydrocracking the low sulfur gas oil product in a second hydrocracking reaction zone at a conversion of at least 20% to form at least one low sulfur fuel product; and
- e) hydrotreating the light gas oil fraction.

1                    DETAILED DESCRIPTION OF THE PREFERRED EMBODIMENTS  
2                    DEFINITIONS

3  
4    For the purposes of this specification, the term "middle distillates" as used  
5    herein is to be taken as a reference to hydrocarbons or hydrocarbon mixtures  
6    having a boiling point or boiling point range substantially corresponding to that  
7    of the kerosene and diesel fractions obtained during the conventional  
8    atmospheric distillation of crude oil feed. The term "light gas oil" (LGO) as  
9    used herein is to be taken as a reference to hydrocarbons or hydrocarbon  
10   mixtures which are isolated as distillate streams obtained during the  
11   conventional atmospheric distillation of a refinery stream, a petroleum stream  
12   or a crude oil stream. The term "vacuum gas oil" (VGO) as used herein is to  
13   be taken as a reference to hydrocarbons or hydrocarbon mixtures which are  
14   isolated as distillate streams obtained during the conventional vacuum  
15   distillation of a refinery stream, a petroleum stream or a crude oil stream. The  
16   term "naphtha" as used herein is a reference to hydrocarbons or hydrocarbon  
17   mixtures having a boiling point or boiling point range substantially  
18   corresponding to that of the naphtha (sometimes referred to as the gasoline)  
19   fractions obtained during the conventional atmospheric distillation of crude oil  
20   feed. In such a distillation, the following fractions are isolated from the crude  
21   oil feed: one or more naphtha fractions boiling in the range of from 30 to  
22   220°C, one or more kerosene fractions boiling in the range of from 120 to  
23   300°C and one or more diesel fractions boiling in the range of from 170 to  
24   370°C. The boiling point ranges of the various product fractions isolated in  
25   any particular refinery will vary with such factors as the characteristics of the  
26   crude oil source, refinery local markets, product prices, etc. Reference is  
27   made to ASTM standards D-975 and D-3699-83 for further details on  
28   kerosene and diesel fuel properties. The term "hydrocarbon fuel" is to be  
29   taken as a reference to either one or a mixture of naphtha and middle  
30   distillates. Unless otherwise specified, all distillation temperatures listed  
31   herein refer to normal boiling point and normal boiling range temperatures.  
32   By "normal" is meant a boiling point or boiling range based on a distillation at  
33   one atmosphere pressure, such as that determined in a D1160 distillation.

1 The term "hydrotreating" as used herein refers to a catalyst process wherein a  
2 suitable hydrocarbon-based feed stream is contacted with a hydrogen-  
3 containing treat gas in the presence of suitable catalysts for removing  
4 heteroatoms, such as sulfur and nitrogen and for some hydrogenation of  
5 aromatics.

6  
7 The term "desulfurization" as used herein refers to a catalyst process wherein  
8 a suitable hydrocarbon-based feed stream is contacted with a  
9 hydrogen-containing treat gas in the presence of suitable catalysts for  
10 removing heteroatoms such as sulfur atoms from the feed stream.

11  
12 The term "hydrocracking" as used herein refers to a catalyst process wherein  
13 a suitable hydrocarbon-based feed stream is contacted with a hydrogen-  
14 containing treat gas in the presence of suitable catalysts for reducing the  
15 boiling point and the average molecular weight of the feed stream.

#### 16 17 CRUDE DESULFURIZATION UNIT

18  
19 The crude oil feed to the present process is generally a whole crude which  
20 has not been substantially separated into individual fractions. Removing  
21 volatile gases and light liquids (including C<sub>1</sub> to C<sub>4</sub> hydrocarbons) prior to  
22 introducing the crude oil feed to the crude desulfurization unit is generally  
23 preferred. The crude oil feed is also treated in a desalting unit prior to  
24 desulfurization. The full benefits of the practice of the invention are equally  
25 realized if a naphtha fraction is removed from the crude oil feed prior to  
26 treating in the crude desulfurization unit.



FIGURE 1  
REACTOR CONFIGURATION

Referring now to Fig. 1, a crude oil feed **02** is passed to a crude desulfurization unit **04** in combination with a hydrogen rich stream **44** for hydrodesulfurizing the crude oil feed. Crude desulfurization unit **04** comprises one or more reaction zones, each of which contains one or more catalyst beds. The crude desulfurization unit removes a substantial portion of the contaminants present in the crude oil feed, including metals, sulfur, nitrogen and Conradson carbon. Catalysts provided in crude desulfurization unit **04** for removing these contaminants may include a single catalyst or a layered catalyst system comprising multiple catalysts present in one or more reactors. When using a reaction train comprising more than one reactor in series operation, a major portion, if not all, of the liquid product from each reactor (except the last reactor vessel in the reaction train) is passed to a next reactor for additional processing. In the layered catalyst system, catalysts are pre-selected for their intended specific use, whether it be demetallation, or sulfur and nitrogen removal, or asphaltene and Conradson carbon removal, or mild conversion. Different catalyst layers may also be selected to facilitate the desulfurization of various boiling point fractions present in the crude oil feed, including naphtha fractions, middle distillate fractions, vacuum gas oil fractions and/or residuum fractions.

DESULFURIZATION UNIT CATALYST

Catalysts for use in the crude desulfurization unit **04** are generally composed of a hydrogenation component, selected from Group VIb (preferably molybdenum and/or tungsten, more preferably molybdenum) and Group VIII (preferably cobalt and/or nickel) of the Periodic Table, or a mixture thereof, all supported on an alumina support. Phosphorous (Group Va) oxide is optionally present as an active ingredient. A typical desulfurization catalyst contains from 3 to 35 wt% hydrogenation components, with an alumina binder.

1 The catalyst pellets range in size from 1/32 inch to 1/8 inch. A spherical,  
2 extruded, trilobate or quadrilobate shape is preferred. In general, the crude oil  
3 feed passing through the desulfurization unit contacts first a catalyst  
4 preselected for metals removal, though some sulfur, nitrogen and aromatic  
5 removal will also occur. Subsequent catalyst layers are preselected for sulfur  
6 and nitrogen removal, though they would also be expected to catalyze the  
7 removal of metals and/or cracking reactions.

8  
9 Catalyst layer(s) preselected for demetallization comprise catalyst(s) having  
10 an average pore size ranging from 125 to 225 Å and a pore volume ranging  
11 from 0.5 - 1.1 cm<sup>3</sup>/g. Catalyst layer(s) preselected for  
12 denitrification/desulfurization comprise catalyst(s) having an average pore  
13 size ranging from 100 to 190 Å with a pore volume of 0.5 - 1.1 cm<sup>3</sup>/g.  
14 U.S. Patent No. 4,90,243 describes a hydrotreating catalyst having a pore  
15 size of at least about 60 Å, and preferably from about 75 Å to about 120 Å. A  
16 demetallation catalyst useful for the present process is described, for  
17 example, in U.S. Patent No. 4,976,848, the entire disclosure of which is  
18 incorporated herein by reference for all purposes. Likewise, catalysts useful  
19 for desulfurization of heavy streams are described, for example, in  
20 U.S. Patent No. 5,215,955 and U.S. Patent No. 5,177,047, the entire  
21 disclosures of which is incorporated herein by reference for all purposes.  
22 Catalysts useful for desulfurization of middle distillate, vacuum gas oil streams  
23 and naphtha streams are described, for example, in U.S. Patent  
24 No. 4,990,243 the entire disclosures of which are incorporated herein by  
25 reference for all purposes.

#### 26 27 REACTION CONDITIONS

28  
29 It is desirable that the crude desulfurization unit **04** be controlled to maintain  
30 the product sulfur at a specified maximum concentration. For example, when  
31 the product sulfur is maintained at less than 1 wt% based on feed, and  
32 preferably less than 0.75 wt% based on feed, reaction conditions in the crude  
33 desulfurization unit **04** include a reaction temperature between about 315°C  
34 and 440°C (600°F – 825°F), pressures from 6.9 MPa to about 20.7 MPa

1 (1000 - 3000 psi), and a feed rate (vol oil/vol cat hr) from 0.1 to about 20 hr<sup>-1</sup>.  
2 Hydrogen circulation rate are general in the range from about 303 std liter  
3 H<sub>2</sub>/kg oil to 758 std liters H<sub>2</sub>/kg oil (2000-5000) standard cubic feet per barrel).  
4

#### 5 DESULFURIZED CRUDE OIL PROPERTIES

6

7 The crude oil desulfurization process removes greater than 25% w/w,  
8 preferably greater than 50% w/w of the sulfur present in the crude oil feed **02**.  
9 The preferred desulfurized crude oil **06** typically has a sulfur content of less  
10 than 1 wt%, preferably less than 0.75 wt%, still more preferably less than  
11 0.5 wt%.

#### 12 DESULFURIZED CRUDE DISTILLATION

13

14 Unreacted hydrogen isolated from crude desulfurization unit **04** is separated  
15 from desulfurized crude oil **06** in one or more flash zones **08** (e.g. a  
16 desulfurization unit high pressure separator) and the resultant desulfurized  
17 liquid **10** is passed to crude fractionator **12** for fractionation to produce at least  
18 a light gas oil fraction **20**, a vacuum gas oil fraction **18** and a residuum  
19 fraction **16**. Crude fractionator **12** is a single or multiple column fractionation  
20 system, and preferably a two column or stage fractionator. One example two-  
21 stage fractionator comprises an atmospheric distillation column operated  
22 substantially at or slightly above atmospheric pressure, and a vacuum  
23 distillation column operated at sub-atmospheric pressure. Such distillation  
24 column systems are well known. In a preferred process of the invention,  
25 desulfurized liquid **10** is passed from flash separation zone(s) **08** directly to  
26 crude fractionator **12** without cooling desulfurized liquid **10** beyond that  
27 required for the distillation in crude fractionator **12**. The temperature of  
28 stream **10** passing from **8** to **12** is preferably maintained at a temperature of at  
29 least 250°F, and preferably of at least 600°F. In the embodiment illustrated in  
30 Fig. 1, all of the desulfurized crude oil, absent light gases, are passed to crude  
31 fractionator **12** for fractionation.

## HYDROCRACKING UNIT

The vacuum gas oil fraction **18** from the crude fractionator **12** is passed to the hydrocracking unit **54**, preferably directly, without tankage and with minimal heat removal, for further processing to produce low sulfur and low aromatic hydrocarbon fuels. The hydrocracking unit **54** contains catalyst selected for further removal of sulfur and nitrogen compounds, for saturation and removal of aromatic compounds, and for cracking for molecular weight reduction. For the present invention, conversion is generally related to a reference temperature, such as, for example, the minimum boiling point temperature of the hydrocracker feedstock. The extent of conversion relates to the percentage of feed boiling above the reference temperature which is converted during hydrocracking into hydrocrackate boiling below the reference temperature. Where the reference temperature is selected to be, e.g. 370°C (700°F), overall conversion during hydrocracking in hydrocracking unit **54** is typically greater than 10%, and preferably greater than 20%.

## 2ND STAGE PRODUCT

Effluent from hydrocracking unit **54** is separated in one or more flash separation units **28** (e.g. hydrocracker separation unit) to isolate at least a hydrocracked liquid product **62**, which is passed to product fractionator **30** for fractionation. In the preferred process, recycle H<sub>2</sub> stream **56** is separated from hydrocracked effluent **52** for recycle to various units in the integrated process, and the remaining liquid **62** is passed to a product fractionator **30** for isolating fuel product(s). The purity of recycle H<sub>2</sub> stream **56** will generally be maintained at greater than 75 mole% hydrogen. In order to maintain energy efficiency, hydrocracked liquid product **62** is passed to fractionator **30** without substantial cooling of **62**. At least one fuel product, **40**, is isolated from product fractionator **30**.

1                                    NAPHTHA PRODUCT

2  
3    Light gas oil **20** is isolated from crude fractionator **12**. This stream may be  
4    blended into a gasoline pool without further processing if desired, particularly  
5    if the sulfur level of light gas oil **20** is below 300 ppm, and preferably below  
6    100 ppm. Alternatively, light gas oil **20** is hydrotreated in hydrotreating  
7    reaction zone **58** to reduce sulfur levels to below 100 ppm, preferably below  
8    50 ppm, and more preferably below 15 ppm. Stream **60** is isolated as  
9    desirably low sulfur naphtha.

10                                   FIGURE 2  
11                                   CRUDE OIL DESULFURIZATION

12  
13    In the preferred embodiment illustrated in Fig. 2, crude oil feed **02** is passed to  
14    crude desulfurization unit **04** for removing contaminants, e.g. one or more of  
15    sulfur, nitrogen, asphaltenes, Conradson carbon, from the crude oil feed **02**.  
16    As described above with respect to Fig. 1, desulfurized crude oil **06** is treated  
17    in one or more flash zones **08** to remove unreacted hydrogen and light  
18    hydrocarbon products **14**. The desulfurized liquid **10** from the flash zone(s)  
19    **08** is then passed to a crude fractionator **12**. In a preferred process of the  
20    invention, desulfurized liquid **10** is passed from flash separation zone(s) **08**  
21    directly to crude fractionator **12** without cooling desulfurized liquid **10** beyond  
22    that required for the distillation in crude fractionator **12**. The temperature of  
23    stream **10** passing from **8** to **12** is preferably maintained at a temperature of at  
24    least 250°F, and preferably of at least 300°F. At least residuum fraction **16**,  
25    vacuum gas oil **18**, and light gas oil **20** are isolated from crude fractionator **12**.

26  
27                                   DESULFURIZED PRODUCT DISTILLATION

28  
29    Fractionation zone **12** may be a single distillation column, or multiple  
30    distillation columns, each positioned in serial flow with respect to the other. In  
31    a preferred embodiment of the process, the desulfurized liquid **10** is  
32    fractionated in fractionation zone **12** which comprises at least one distillation  
33    column (not shown) which is operated substantially at or slightly above  
34    atmospheric pressure (i.e. an atmospheric distillation column) and at least one

1 distillation column (not shown) which is operated at sub-atmospheric pressure  
2 (i.e. a vacuum distillation column). Such distillation columns are well known in  
3 the art. Desulfurized liquid **10** is passed to the atmospheric distillation column  
4 to produce at least naphtha stream **20** and an atmospheric residuum, which is  
5 further fractionated in the vacuum distillation column. A vacuum gas oil **18** is  
6 isolated as a distillate fraction from the vacuum distillation column, and  
7 vacuum residuum stream **16** is isolated as a bottoms fraction from the  
8 vacuum distillation column.

9  
10 The vacuum gas oil **18** is passed directly to hydrocracker unit hydrocracking  
11 unit **54** for conversion to lower molecular weight products and for reduction in  
12 sulfur, nitrogen and/or aromatic content. As shown in the preferred  
13 embodiment illustrated in Fig. 2, the hydroconversion step involves at least  
14 two reaction vessels, first hydrocracker stage **22** and second hydrocracker  
15 stage **26**. The hydrocracking process is especially useful in the production of  
16 middle distillate fractions boiling in the range of about 250°-700° F.  
17 (121°-371° C.) as determined by the appropriate ASTM test procedure. The  
18 hydrocracking process involves conversion of a petroleum feedstock by, for  
19 example, molecular weight reduction via cracking, hydrogenation of olefins  
20 and aromatics, and removal of nitrogen, sulfur and other heteroatoms. The  
21 process may be controlled to a certain cracking conversion or to a desired  
22 product sulfur level or nitrogen level or both. Conversion is generally related  
23 to a reference temperature, such as, for example, the minimum boiling point  
24 temperature of the hydrocracker feedstock. The extent of conversion relates  
25 to the percentage of feed boiling above the reference temperature which is  
26 converted during hydrocracking into hydrocrackate boiling below the  
27 reference temperature.

## HYDROGEN RECOVERY

The hydrogen stream **14** isolated from flash separation zone **08** may be further purified in, for example, an amine scrubber **46** to remove some or all of the  $H_2S$  and  $NH_3$  gases. Following compression, the purified hydrogen is passed to the first hydrocracker stage **22** and the second hydrocracker stage **26**.

### 1ST STAGE

Reaction in first hydrocracker stage **22** is maintained at conditions sufficient to further remove nitrogen and sulfur contaminants from the vacuum gas oil feed **18** and for reducing the aromatic content of the vacuum gas oil feed **18**. These hydrotreating reactions are generally characterized by a low amount of conversion, e.g. less than 20%, preferably less than 15%. In general, it is desirable to lower the nitrogen content of the hydrocarbon feedstock stream to less than 50 parts per million by weight (ppm), preferably less than about 10 ppm and for increased catalyst life to a level of less than 2 ppm or even as low as about 0.1 ppm. Similarly, it is generally desirable to lower the sulfur content of the hydrocarbon feedstock stream to less than about 0.5% by weight percent, preferably less than about 0.1%, and in many cases as low as about 1 ppm.

### 1ST STAGE CONDITIONS

Thus, the one or more reaction zones in first hydrocracker stage **22** are operated at reaction temperatures between  $250^{\circ}C$  and about  $500^{\circ}C$  (  $482 - 932^{\circ}F$  ), pressures from 3.5 MPa to about 34.2 MPa (500 - 3500 psi) , and a feed rate (vol. oil/vol. cat h) from 0.1 to about  $20hr^{-1}$  . Hydrogen circulation rates are in general in the range from about 350-std. liter  $H_2/kg$  oil to 1780  $H_2/kg$  oil (2310 - 11750 standard cubic feet per barrel). Preferred reaction temperatures range from  $340^{\circ}C$  to about  $455^{\circ}C$  (  $644 - 851^{\circ}F$  ). Preferred total reaction pressures range from 7.0 MPa to about 20.7 MPa (1000 - 3000 psi ).

## 1ST STAGE CATALYST

Catalysts useful in first hydrocracker stage **22** generally contain at least one Group VIb metal (e.g. molybdenum) and at least one Group VIII metal (e.g. nickel or cobalt) on an alumina support. A phosphorous oxide component and a cracking component, such as silica-alumina and/or a zeolite, may also be present. A layered catalyst system may also be used, e.g. the layered catalyst system taught in U.S. Patent No. 4,990,243, which is incorporated herein by reference for all purposes. The catalyst selected for use in first hydrocracker stage **22** will generally have a pore volume in the range of 0.5 to 1.2 cm<sup>3</sup>/g, with an average pore diameter of between 100 Å and 180 Å, and a surface area 120 and 400 m<sup>2</sup>/g, wherein at least 60% of the pores have a pore diameter of more than 100 Å. The first stage catalyst could also be a layered system of hydrotreating and hydrocracking catalysts. The preferred catalyst for first hydrocracker stage **22** comprises a nickel molybdenum or cobalt molybdenum hydrogenation component and a silica-alumina component with an alumina binder.

## HOT H<sub>2</sub> STRIPPER

The effluent **48** from the first hydrocracking stage **22** contains unreacted hydrogen, gaseous and liquid products. Hydrogen isolated from effluent **48** contains H<sub>2</sub>S and NH<sub>3</sub>. In conventional processes, such hydrogen is purified prior to use as recycle to the first hydrocracking stage or as H<sub>2</sub> feed to the second hydrocracking stage. The present process is based on the realization that hydrogen isolated from effluent **48** is suitable for use as H<sub>2</sub> feed to the crude desulfurization unit **04**, without extensive purification. The use of hydrogen in this way is facilitated by passing effluent **48** to hot hydrogen stripper **24** for removing light gases contained therein, including hydrogen and light hydrocarbon gases, using heated hydrogen **36**. Typically, hot hydrogen stripper **24** is operated at temperatures preferably between 260°C and 399°C (500°F and 750°F). Hydrogen-rich stream **44**, which is isolated from hot hydrogen stripper **24**, is combined with crude oil feed **02**, preferably with no further purification, for desulfurizing crude oil feed **02** in crude desulfurization



1 unit **04**. Stripped effluent **50** isolated from hot hydrogen stripper **24** is passed  
2 to second hydrocracker stage **26** for further upgrading. In a preferred  
3 embodiment of the process, effluent **48** passes directly from reaction zone **22**  
4 to a single stage **24** for hot hydrogen stripping. Stripped effluent **48** is then  
5 passed directly as a heated liquid, with no cooling beyond the normal minimal  
6 cooling associated with movement through the pipes connecting the various  
7 processing units, to second hydrocracker stage **26** for further reaction.

## 8 9 2ND STAGE

10  
11 Second hydrocracker stage **26** is a hydrocracking stage, operated at  
12 hydrocracking conditions and with a catalyst(s) suitable for molecular weight  
13 reduction, with additional sulfur, nitrogen and aromatics removal. Conditions  
14 in second hydrocracker stage **26** are suitable for per pass conversions of up  
15 to 90%. Indeed, operating second hydrocracker stage **26** in extinction recycle  
16 mode, with partially reacted product being recycled until all have been  
17 cracked, is also within the scope of the present process.

## 18 19 2ND STAGE CONDITIONS

20  
21 The hydrocracking conditions used in the hydrocracker will range from 250°C  
22 to about 500°C (482 - 932°F), pressures from about 3.5 MPa to about  
23 24.2 MPa (500 - 3500 psi), and a feed rate (vol. Oil/ vol. cat h) from 0.1 to  
24 about 20 hr<sup>-1</sup>. Hydrogen circulation rates are generally in the range from  
25 about 350 std liters H<sub>2</sub>/kg oil to 1780 std liters H<sub>2</sub>/kg oil (2310 - 11750  
26 standard cubic feet per barrel). Preferred total reaction pressures range from  
27 7.0 MPa to about 20.7 MPa (1000 - 3000psi). Second hydrocracker stage **26**  
28 is operated at temperatures of greater than 650°F and pressures between  
29 about 1000 psig and 3500 psig, preferably between 1500 psig and 2500 psig  
30 hydrogen pressure.

## 2ND STAGE CATALYST

The catalyst used in the second hydrocracking stage **26** is a conventional hydrocracking catalyst of the type used to carry out hydroconversion reactions to produce transportation fuels. First hydrocracker stage **22** and second hydrocracker stage **26** can contain one or more catalyst in more than one reaction zone. If more than one distinct catalyst is present in either or the reaction zones, they may either be blended or be present as distinct layers. Layered catalyst systems are taught, for example, in U.S. Patent No. 4990243. Hydrocracking catalyst useful for second hydrocracker stage **26** are well known. In general, the hydrocracking catalyst comprises a cracking component and a hydrogenation component on an oxide support material or binder. The cracking component may include an amorphous cracking component and/or a zeolite, such as a y-type zeolite, and ultrastable Y type zeolite, or a dealuminated zeolite. Particularly preferred catalytic cracking catalysts are those containing at least one zeolite which is normally mixed with a suitable matrix such as alumina, silica or silica-alumina. A suitable amorphous cracking component is silica-alumina. The preferred amorphous cracking component is between 10 and 90 weight percent silica, preferably between 15 and 65 weight percent silica, the remainder being alumina. A cracking component containing in the range from about 10% to about 80% by weight of the Y-type zeolite and from about 90% to about 20% by weight of the amorphous cracking component is preferred. Still more preferred is a cracking component containing in the range from about 15% by weight to about 50% by weight of the Y-type zeolite, the remainder being the amorphous cracking component. Also, so-called x-ray amorphous zeolites (i.e., zeolites having crystallite sizes too small to be detected by standard x-ray techniques) can be suitably applied as cracking components. Hydrogenation components suitable for the hydrocracking and/or hydrotreating catalysts which are used in the present integrated process include those which are comprised of at least one Group VIII (IUPAC Notation) metal, preferably iron, cobalt and nickel, more preferably cobalt and/or nickel and at least one Group VI (IUPAC Notation) metal, preferably molybdenum and tungsten, on a high surface area support

1 material, preferably alumina. Other suitable catalysts include zeolitic  
2 catalysts, as well as noble metal catalysts where the noble metal is selected  
3 from palladium and platinum. It is within the scope of the present invention  
4 that more than one type of catalyst be used in the same reaction vessel. The  
5 Group VIII metal is typically present in an amount ranging from about 2 to  
6 about 20 weight percent. The Group VI metal will typically be present in an  
7 amount ranging from about 1 to about 25 weight percent. The hydrogenation  
8 components in the catalyst may be in the oxidic and/or the sulfidic form. If a  
9 combination of at least a Group VI and a Group VIII metal component is  
10 present as (mixed) oxides, it will be subjected a sulfiding treatment prior to  
11 proper use in hydrotreating or hydrocracking. Suitably, the catalyst comprises  
12 one or more components of nickel an/or cobalt and one or more components  
13 of molybdenum and/or tungsten or one or more components of platinum  
14 and/or palladium. Catalysts containing nickel and molybdenum, nickel and  
15 tungsten, platinum and/or palladium are particularly preferred.  
16  
17 The effective diameter of the zeolite catalyst particles are in the range of from  
18 about 1/32 inch to about 1/4 inch, preferably from about 1/20 inch to about 1/8  
19 inch. The catalyst particles may have any shape known to be useful for  
20 catalytic materials, including spheres, cylinders, fluted cylinders, prills,  
21 granules and the like. For non-spherical shapes, the effective diameter can  
22 be taken as the diameter of a representative cross section of the catalyst  
23 particles. The catalyst particles will further have a surface area in the range of  
24 from about 50 to about 500 m<sup>2</sup>/g.

#### 25 26 LAYERED HYDROCRACKING ZONE FOR LIGHT GAS OIL 27 HYDROTREATING 28

29 In Fig. 1, a light gas oil stream **20** isolated from the desulfurized liquid **10** is  
30 hydrotreated in **58** to remove sulfur and/or aromatics in preparation of a low  
31 sulfur, low aromatic fuel product **60**. In a separate preferred embodiment  
32 illustrated in Fig. 2, the hydrotreating catalyst useful for hydrotreating light gas  
33 oil stream **20** is layered at or near the bottom of second hydrocracker stage  
34 **26**. Thus, second hydrocracker stage **26** includes a layered catalyst system,

1 with catalysts typically used for hydrocracking near the feed inlet to second  
2 hydrocracker stage **26** and one or more layers of catalyst typically used for  
3 hydrotreating near the product effluent outlet of second hydrocracker stage  
4 **26**. The amount of hydrotreating catalyst in second hydrocracker stage **26** is  
5 generally smaller than the amount of hydrocracking catalyst included in  
6 second hydrocracker stage **26**. In including the hydrotreating catalyst as a  
7 layer in an otherwise hydrocracking reaction mode, it is expected that the  
8 effluent from the catalyst layers for hydrocracking, having reacted at  
9 hydrocracking conditions in second hydrocracker stage **26**, would not be  
10 modified to any significant extent in the layer of hydrotreating catalyst in  
11 second hydrocracker stage **26**. However, the unreacted hydrogen in the  
12 reacting stream passing from the bed(s) of hydrocracking catalyst are  
13 available for further reaction without additional heating, pressurization and/or  
14 purification. Thus, light gas oil stream **20** stream, which is essentially fuel  
15 boiling range material, but with higher amounts of sulfur, nitrogen and/or  
16 aromatics than is permitted for current fuels, is passed to the portion of  
17 second hydrocracker stage **26** which contains the layer(s) of hydrotreating  
18 catalyst. Bypassing the hydrocracking catalyst beds reduces the amount of  
19 undesirable cracking of light gas oil **20** stream. Furthermore, reaction of light  
20 gas oil stream **20** in combination with the effluent from the layers of  
21 hydrocracking catalyst of second hydrocracker stage **26** serves to remove  
22 additional contaminants from light gas oil stream **20** without molecular weight  
23 reduction and without added hydrogen beyond that potentially required to  
24 quench exothermic heat release from the layers of hydrotreating catalyst in  
25 second hydrocracker stage **26**. The reaction conditions for hydrotreating the  
26 naphtha stream in the second hydrocracker stage is expected to be the same  
27 as reaction conditions for hydrocracking in that stage. The blend of fuels  
28 produced in the various catalyst layers of second hydrocracker stage **26** is  
29 separated in product fractionator **30**. At least one fuel stream, shown as **40** in  
30 Fig. 2, is isolated from product fractionator **30**.

## 2ND STAGE PRODUCT

1  
2  
3 Effluent **52** from second hydrocracker stage **26** is separated in hydrocracker  
4 flash separation zone(s) **28** to isolate at least a recycle hydrogen stream **42**  
5 and a hydrocracked liquid product **62**, which is passed to product fractionator  
6 **30** for fractionation. At least one low sulfur fuel product, **40**, is isolated from  
7 product fractionator **30**. However, it is expected that a full range of fuel  
8 products, including low sulfur naphtha, low sulfur kerosene and low sulfur  
9 diesel would desirably be isolated in the process. Stream **56** is combined with  
10 fresh hydrogen **32** and with isolated hydrogen stream **14** as hydrogen feed to  
11 first hydrocracker stage **22**, to hot hydrogen stripper **24** to second  
12 hydrocracker stage **26**. Incompletely reacted products from second  
13 hydrocracker stage **26** are recycled via **42** to second hydrocracker stage **26**.